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(54) Title: NEW PROCESS FOR BUTYL RUBBER PRODUCTION			
<p>(57) Abstract</p> <p>This invention relates to a continuous slurry process for the production of Butyl Rubber (IIR) and to the devices thereof. The new process allows substantial energy and investment saving vis-à-vis conventional processes. In the new process the polymerization is performed at -50 °C instead of -100 °C, by using special catalysts. The polymerization heat is removed by allowing the reaction medium to boil at reduced pressure, instead of exchanging it through the reaction walls. The polymerization reactor is a tri-phasic reactor, i.e. it contains a polymer dispersed phase, a liquid phase (composed of monomers and methylchloride), and a vapor phase. The reactor is designed in order to: i) homogenize the liquid phase in every point of the reactor; ii) convey (to "skim") the formed polymer particles across the reactor toward the slurry outlet; iii) separate the vapor phase on the upper part of the reactor (reactor dome) where the vapors are taken away by a vapor outlet. The slurry discharged by the reactor is taken away by a vertical extruder adjacent and connected to the reactor by a slit. The extruder screw conveys and concentrates by compression the slurry, whereas the liquid phase flows back to the reactor. In this invention, the combination of the abovesaid critical characteristics allows the production energy of IIR to be abated by more than 20% and the investment by more than 30 %. The process scheme is shown in fig. 1.</p>			

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DESCRIPTION

NEW PROCESS FOR BUTYL RUBBER PRODUCTION

This invention relates to a new process for production of isobutene polymers and co-polymers, and to the devices thereof.

In particular, the invention relates to the 5 production of high MW isobutene-isoprene copolymers, known as Butyl Rubber or "IIR".

IIR is known for about 50 years (U.S. 2356128 and U.S. 2356130). It finds application as a low gas diffusion and oxidation resistant rubber. IIR is commercially produced in the presence of cationic catalysts, also known as Friedel-Craft catalysts, at extremely low temperatures (e.g. 173 K°), in a reaction medium of methylchloride. Unlike to its monomers, IIR is not soluble in methylchloride, so that a polymer slurry is formed. Details on this technology are available in the literature (C. Whitby Synthetic Rubber, Chapt.24 J. Wiley & Sons and Kirk Othmer Encyclopedia of Chem. Technol., Butyl Rubber).

Even if the present IIR processes are still valid, 20 they show important drawbacks. These drawbacks can be summarized as follows:

- i) the energy consumption of the process is very high
- ii) the plant investment cost is also very high
- 25 iii) the quality constancy control is difficult

In more detail :

i) The high energy consumption is due to a series of cooling cycles for supplying the process with the necessary frigories at 173 K°. As an order of magnitude :

5 tude the energy cost of cooling cycles can be as high as 75% of the total energy requirement.

10 ii) The investment cost of a IIR plant is also very high. Beside the cost of the mentioned cooling cycles, which can be as high as 35% of the total investment, the polymerization reactors are very expensive. In fact their design is very complex and they are built with expensive cryogenic materials. The investment cost is exacerbated by the fact that every 2-30 reactors in operation a supplementary stand-by reactor is needed, 15 since the internal heat exchangers fouling imposes a rather frequent reactor shut-down for cleaning.

20 iii) The IIR quality depends on its MW, and in turn MW depends on polymerization temperature. The above mentioned reactor fouling is conducive to a temperature increase inside the reactor. Consequently, the operating life of the reactor lasts for a few ten hours. This fact has an impact on IIR quality constancy.

25 In order to overcome the aforementioned drawbacks, several solutions have been proposed and patented. However these solutions were not solving the basic problems of IIR process, or were not practically applicable.

30 Among the relevant proposed solutions, the use of new catalyst systems, called by some researchers "syncatalysts" has been put forward about 20 years ago.

35 The syncatalysts allowed in principle to obtain high MW IIR even at a temperature as high as 223K°, vis a vis the 173K° of the conventional processes. However, the practical use of syncatalysts has been ruled out by the worsening of the already critical problem of reactor

fowling.

More recently a new IIR process (U.S. 4714747) has been patented in which syncatalysts are used without meeting heat exchange problems due to reactor fowling.

According to U.S. 4714747, monomers and reaction medium are fed into a double screw, self cleaning extruder. Inside this extruder the reactant fluids followed a peculiar fluidynamics known as "plug flow". The boiling of reaction medium inside the extruder was avoiding the need of a heat exchange through the reactor walls. However, as point out in U.S. 4714747, the double screw extruder was compacting the polymer in a "viscous and sticky mass" which hampered the vapor separation and, consequently, the reactor temperature control.

In conclusion, also this solution was difficult to put in pratice.

All the aforementioned drawbacks are overcome by the process and by the devices of this invention, in which are contemporarily present all the following critical characteristics:

1) The polymerization is performed at temperatures between 208 and 288K°, in the presence of syncatalysts, whose two (or more) components are fed directly and separately to the polymerization reactor.

2) The polymerization heat is removed by allowing the reaction medium to boil inside the reactor, and maintaining the reactor content as a low viscosity polymer slurry.

3) The polymerization reactor is designed in order to keep the reactor content as a low viscosity slurry. A coalescence of polymer particles in bigger

and bigger crumbs is avoided, which is conducive to the "viscous and sticky mass" shown in U.S. 4714747. In the reactor of the present invention, the vapors produced by the exothermic polymerization reaction can easily 5 escape from the polymer slurry, owing to its low visco sity, and collect in the reactor dome.

4) The polymerization reactor is designed in order to fulfill other important duties : one is that of homogenizing the liquid phase in every point of the 10 reactor condensed phase, in order to maintain a constant composition of reactants. The second is that of conveying the polymer phase across the reactor, toward the polymer slurry outlet. The third is that of concentrating the polymer slurry discharged by the reactor. 15 The last two duties, which are somewhat recalling certain dairy processes, have been dubbed with the word "skimming".

5) The skimmed slurry flowing from the reactor outlet is concentrated further on in a vertical extruder 20 connected with the reactor. This vertical extruder, referred to as "discharge screw", conveys and presses the solid polymer contained in the slurry upward and "squeezes" the liquid downward (i.e. backward) toward the reactor. In this way, valuable catalyst components, soluble 25 in the reaction medium, are re-fluxed to the reactor for a further utilization.

The details of the process of the present invention and of the devices for its practical application are dealt with herebelow.

30 Polymerization reactor

The reactor is a vertical cylindrical vessel,

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equipped with a special stirrer. In fig. 2 a vertical section is shown. The reactor wall is indicated by 1; 2 and 3 are respectively the reactor ceiling and bottom, 4 is the stirrer shaft, 5 is a standing shaft fastened to the reactor bottom, 6 is one of the vertical paddles, moved by shaft 4 to which they are connected by the cross bar 7. The rotating bars 8 intermesh with standing crossbars 9. 10 is an extruder barrel fastened to the reactor, to which it is connected via slit 12. 10 This extruder, equipped with its screw 11, is hereinafter referred to as "discharge screw".

An horizontal reactor section is also shown in fig. 2. The section is along a standing bar 9. In this figure, three paddles 6, one bar 9 and the slit 12 are 15 shown. The three paddles 6 have a peculiar shape and orientation. When they rotate according to the direction indicated by an arrow, the clearance between a point on the reactor wall and the paddle decreases till to a very small clearance.

20 The reactor is equipped with on inlet for the monomers/polymerization medium mixture, and of two inlets for feeding the catalyst components. The reactor is also equipped with a vapor outlet, placed in the reactor ceiling.

25 The following description of the reactor operation in a continuous, steady state operation will better explain the function of every reactor fitting.

30 The reactor operates at constant level of condensed phase (which includes liquid monomers and reaction medium and polymer particles).

IIR particles are continuously formed in the

reactor, and are immediately conveyed by the stirrer across the reactor, toward its peripheric part, where the slurry discharge 12 is placed, with a continuous skimming action.

5 The rather small clearance among moving and standing surfaces of the stirrer/reactor system limits the surfaces fowling, but does not hamper the liquid phase to flow across the clearances. In the way, the liquid phase is homogenized throughout all the reactor.

10 The polymer skimming is performed in the following way: the polymer slurry present in the peripherical part of the reactor is continuously intaken into the rather large clearance between the paddle leading edge and the reactor wall. The liquid phase. 15 can easily flow through the clearance between the paddle rear edge and the reactor wall. On the contrary the polymer particles are "catched" owing to their size. Therefore the slurry concentration in the space between the paddle and the reactor wall increases during the paddles motion. The rear paddle edge squeezes the polymer slurry against the wall for a phenomenon of visco-dynamic pressurization (Z. Tadmor, C. Gogos Principles of Polymer processing Whiley-Interscience Chap. 10), which increases the solid content of the slurry, 20 whereas the liquid is squeezed out. Whenever a paddle 6 passes in front of slit 12, the slurry is pumped through it, toward the discharge screw 11, which carries it up. Therefore, whenever a paddle 6 passes in front of slit 12, the slurry contained in the volume 25 between paddle and wall is cleared out, and is ready 30 to be re-filled by another amount of slurry, coming

from the axial part of the reactor. A step carved in the rear edge of paddles 6 decreases the gap between paddle and wall and increases the skimming efficiency.

At a certain stirrer rotation speed, the 5 polymer discharged by the reactor in unit time increases proportionally to the polymer concentration.

At a certain polymer concentration, the polymer discharged by the reactor in unit time increases proportionally to the stirrer rotation speed. 10

Consequently, the above reactor has a built-in capacity of controlling the polymer concentration inside the reactor. Once a certain value of polymer output has been selected, by feeding the catalyst at a certain rate, the polymer concentration inside the 15 reactor cannot overcome an upper threshold limit, which is imposed by the stirrer rotation speed. In this way, any excessive polymer content inside the reactor is avoided, with the related risks of reactor "blocking" or of excessive mechanical stress applied 20 to the stirrer.

The process

Details on the process of the present invention are given herebelow, with reference to fig. 1.

The polymerization reactor is indicated by 25 number 1. The reactor outlet 12 joins the reactor to the discharge screw 11, which in turn is connected via throttle valve 13, to the double screw devolatilizer 14. Monomers and reaction medium make-up are fed via lines 15, A, B and C to heat exchanger 16, where they 30 are cooled down to polymerization temperature, and therefrom to reactor, via line 17. The two catalyst compo-

nents are separately fed to the reactor, via lines 18A and B. As soon as the catalyst components come in contact with the monomers the polymerization takes place. Polymer particles are produced around the active centers. It has been observed with surprise that a quick withdrawal of polymer particles from the reactor (above indicated as skimming) does not influences neither the polymerization rate nor the polymer yield. The polymerization is performed at temperatures between 208 and 288K°, at pressures correspondent to the vapor pressure of the reacting system, in general between 0,1 and 4 bar. In this way the reactor content is allowed to boil. The vapors are collected in the reactor dome and then, via line 19 to the knockout drum 20, and, via line 21, sent to compressor 22. The compressed vapors are cooled and condensed in heat exchanger 23. The condensate is flashed in flash drum 24, via throttle valve 25, which drops the pressure to the same level as that of the reactor 1. In flash drum 24 a part of the condensate fed to throttle valve 25 is vaporized and recycled to knock-out drum 20, via line 26, and therefrom to compressor 22, via line 21, together with the vapors coming from the reactor, via line 19. A part of the condensate fed to throttle valve 25 is collected in flash drum 24 as a liquid, at a temperature close to that of reactor 1. This liquid is sent to the reactor under level control, via line 27. In a steady state, the vapors produced by reactor 1 are almost completely recycled to it as liquids, at a temperature close to that of the reactor. Summing up, the polymerization heat is removed from the reactor

as latent vaporization heat.

Any non condensable in heat exchanger 23 is purged via an outlet placed in the coldest part of 5 heat exchanger 23 and sent, via line 28, to a vacuum pump (not shown in the figure).

The slurry produced by the reactor is concentrated, as already said, in the discharge screw 11, fed through throttle valve 13 to mechanical devolatilizer 14, where the residual liquids are removed, by 10 combined thermal and mechanical action, as vapors via line 29. For the sake of exemplification the polymer concentration at the reactor outlet can be high as 50% wt and at the discharge screw outlet as high as .70% wt. 15 The produced IIR shows a high MW. Number average MN's between 100000 and 500000 are obtained by the process as above described.

The catalysts

Suitable catalysts for the process of the 20 present invention are the syncatalysts, which have been extensively patented BE 663319, BE 663320, GB Pat. Nos. 1362295, 1407414, 1407415, 1407416, 1407417, 1407418, 1407419, 1407420, 1409337; The composition 25 and the characteristics of said catalysts are broadly described in the mentioned patents and in several publication on this subject. (M. Baccaredda, M. Bruzzone et al Chim. e Ind. (Milano) 55,109 (1973). A characteristic of syncatalysts is the fact that the active centers are obtained by interaction of, at least, 30 two components, in a stoichiometric ratio not lower/higher than three decades, unlike conventional Friedel-Craft catalysts, in which the co-catalyst, or co-initi-

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tiator, is present in a very small amount, sometime non measurable amount. Several syncatalysts are made 5 of just two components: The first one belongs to the class of Aluminium trialkyls or of Aluminium dialkyl monohalides, as for ex. AlEt₃, AlEt₂Cl, AlEt₂Br, AliBu₂Cl etc.

Typical examples of second components of syncatalysts are : halogens, interhalogenic compounds 10 and organic halides which contain at least a halogen atom susceptible to be exchanged with the first component of the catalyst system in the process conditions. Cl₂, Br₂, Iodine chloride, tert-butylchloride, tert-butylbromide, chloranil are non-limitative examples 15 of second component of syncatalysts. The ratio between the two catalyst components is in general indicated by the atomic ratio Al/X, where X is the halogen present in the second component of the catalyst mixture. The 20 Al/X ratio is in general higher than 1 and can be as high as 1000, the preferred ratio ranging from 10 to 300.

The consumption of second component, expressed 25 in mol per mol of polymerized monomer, is in general ranging from 1×10^{-6} and 50×10^{-6} .

EXAMPLE

The flow sheet of the polymerization run 30 of this example is show in fig. 1. The reactants amount, expressed in kg/h, and their temperature, expressed in C°, in steady state conditions are shown in the attached Table. In this Table also the cross-reference numbers with fig. 1 are indicated.

The reactor of the experiment was the one

described in fig. 2. The reactor was insulated with
 5 polystyrene foam. A certain quantity of pure Methyl-
 chloride (MeCl) was fed as a liquid, via line 15 C,
 to the exchanger 16, where it was cooled to -45°C, and
 therefrom to reactor 1, via line 17, up to a selected
 10 level. Isobutene and isopropene were then fed, via
 lines 15 A and B; in the quantity shown in the Table,
 and the two components of the catalyst, via lines
 18 A and B. The two components were Aluminumdiethyl-
 monochloride (DEAC)

15

TABLE

POINT OF ANALYSIS	Polym	MeCl	IC4 ^m	Isop.	T(C°)	P(kg/cm ²)
17 (Feeding)		6.1	14.6	0.4	45	0.34
19 (Vapor from reactor)		20.6	4.4	0.7	45	0.34
27 (Recycle to reactor)		20.6	4.4	0.7	46	0.34
29 (Volatile from desolv.)		6.1	2.4	0.1	100	0.20
30 (Dry polymer)	12.5	-	-	-	100	1.00

Units: kg/h

25

and tert-Butylchloride (TBC). Upon feeding of the two catalyst components, a milky slurry of polymer was immediately formed. The reactor temperature and pressure showed a trend to increase. The pressure increase fostered the intervention of the pressure control placed on the knock-out drum 20, and compressor 22 began to suck the reactor vapors, via line 19 and knock-out drum 20. From this moment onward the pressure in the reactor remained constant at the value of
 30

5 0.34 bar, corresponding to the equilibrium temperature of -45°C in the reactor. The other controls shown in fig. 1 were put into operation. The vapors compressed by compressor 22 were condensed in heat exchanger 23 and the condensate was expanded in vessel 24, kept
10 at the same pressure of the reactor. The condensate was partially vaporized, with a consequent temperature drop, till to a value close to that of the reactor. The cold liquid collected in vessel 24 was sent to the reactor under level control, via line
15 27, whereas the vapors returned to the compressor via line 26 and knock-out drum 24. In this way the vapors produced by the reactor were quantitatively recycled to the same as liquids, at a temperature close to its temperature. The polymer concentration in the reactor increased till to a steady value, at which the formed
20 polymer was quantitatively removed by discharge screw 11. The polymer concentration at throttle valve 13 was about 60%. The polymer was fed, through valve 13 in desolventizer 14, where by combined thermal and mechanical action, the residual liquids were vaporized and recovered via line 29. The IIR free from volatiles was collected as a rubbery extrudate on a conveyor belt 30. The polymer output was about 12,5 kg/h.

30 The formation of polymer in the reactor and its removal by discharge screw 11 were decreasing the monomers concentration in the reaction medium and the polymer mass contained in the reactor. The first effect was detected by a decrease of monomers concentration in the vapors at the reactor out
35

5 let (line 19), controlled by an analyzer which fostered an increase of isobutene feed. In turn, this increase fostered an increase of isoprene feed, through a controller of feed ratio. The second effect was detected by a decrease of liquid level in the reactor,
10 which fostered the intervention of the reactor level controller, which increased the MeCl feed. All fresh monomers and MeCl were cooled to -45 C° in heat exchanger 16.

According to the above flow sheet, the polymerization heat was removed as latent heat of vaporization, and the vapor output from the reactor was dependent on catalyst concentration inside the reactor. Consequently, the feed of one component of the catalyst system was controlled by the vapors output of
20 the reactor. The feed of the second component of the catalyst was kept proportional to the first one by a controller of feed ratio.

CLAIMS

1. A continuons process for the production of high molecular weight isobutene-isoprene copolymers, 5 containing from 0.5 to 10wt-% of combined isoprene, said process being characterized by the contemporary presence of the following critical characteristics in the various process sections :

10 i) polymerization section characterized by the fact that the copolymer is formed, and is maintained during the polymerization course, as a suspension of particles in a liquid phase, said liquid phase being composed of monomers and methylchloride, at a temperature chosen in the range between -65 and +15°C, 15 in the presence of polymerization catalysts obtained by combination of a least two components, the first of which being selected among aluminum trialkyls and aluminum dialkylmonohalides, and the second among halogens, interhalogenic compounds, alkyl- 20 and arylhalides, said satalyst components being fed separately to the polymerization, the polymerization temperature being kept constant by boiling the liquid phase at a pressure between 0.1 and 4,0 bar.

25 ii) discharge section of the polymer slurry from the polymerization section characterized by the fact that said polymer suspension (slurry) is conveyed and compressed upwards, squeezing the liquid phase downwards, so re-fluxing it to the polymerization section.

30 iii) devolatilization section characterized by the fact the concentrated polymer slurry, obtained in the preceding point ii) is transformed into a solid

rubbery polymer, free of volatiles, whereas the liquid phase contained in the slurry is vaporized and recovered by condensation.

2. Devices suitable for the realization of the process of claim 1), i.e.:

5 the vapors obtained by vaporization of the liquid phase contained in the slurry fed to the desolventizer.

3. A process according to claim 1, in which the first component of the catalyst system is chosen among aluminum trialkyls and aluminum dialkylmono-halides.

10 4. A process according to claim 1, in which the second component of the catalyst system is chosen among halogens and alkylhalides.

5. A process according to claim 1, in which the polymer slurry at the desolventization inlet 15 has a solid concentration between 60 and 70wt-%.

6. Butyl rubber with a isoprene content lower than 5wt-% and a number average molecular weight between 50000 and 500000, obtained according to the preceding claims 1 to 5.

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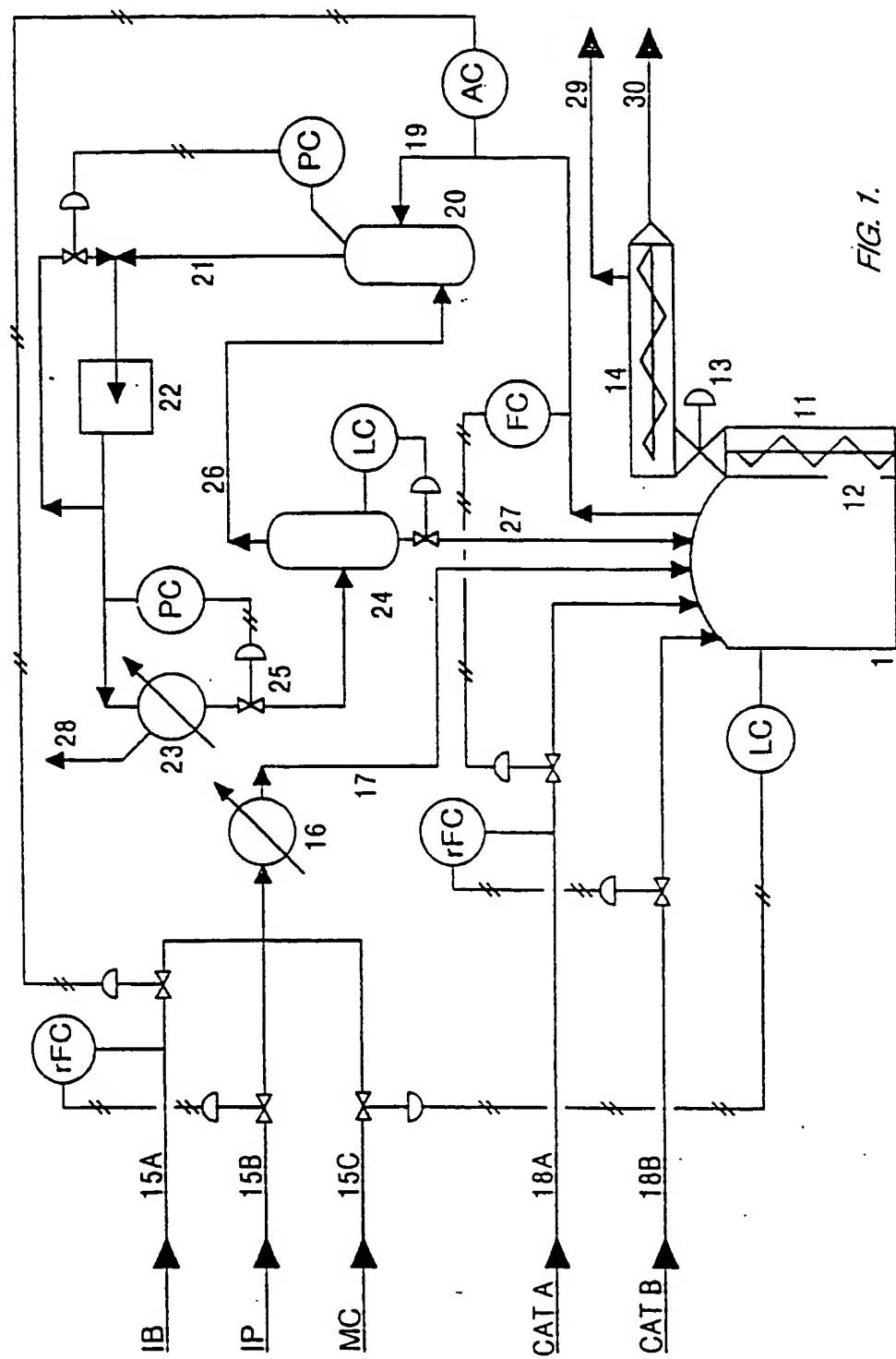


FIG. 1.

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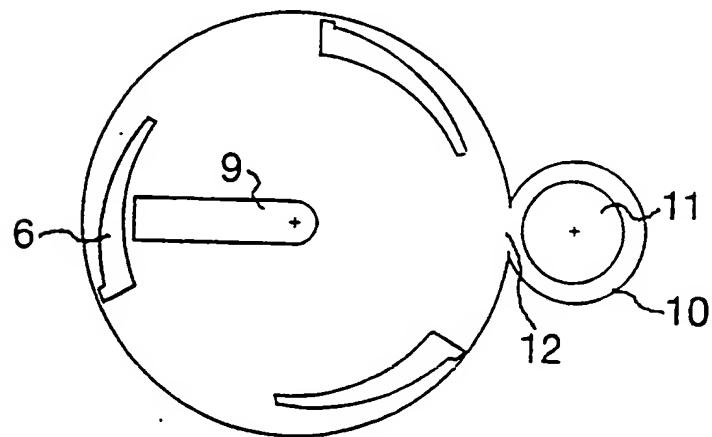
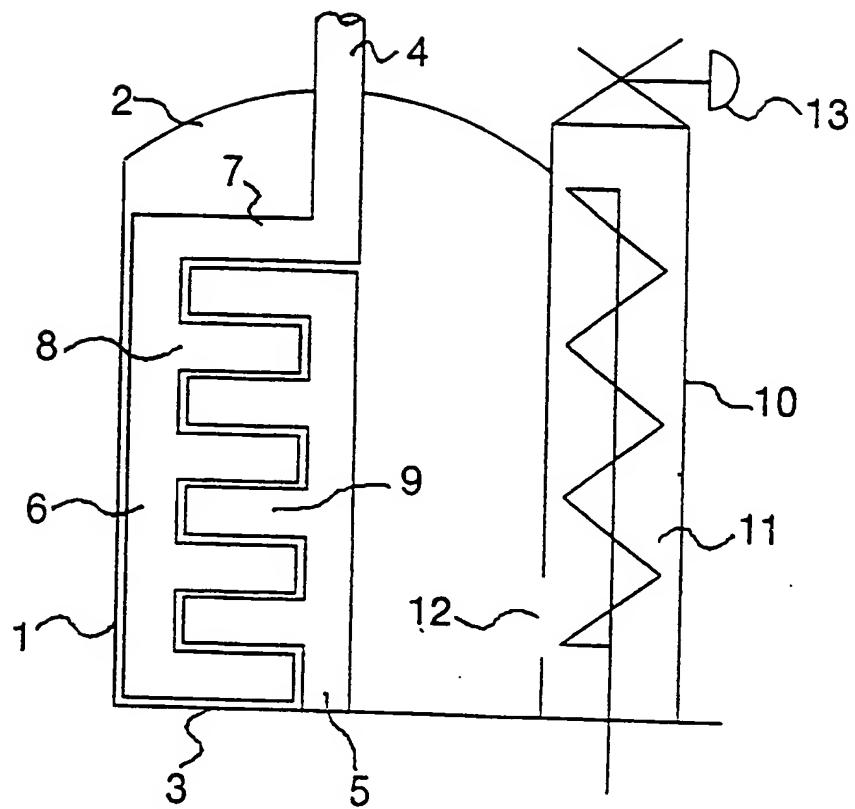


FIG. 2.

INTERNATIONAL SEARCH REPORT

PCT/IT 93/00033

I. CLASSIFICATION OF SUBJECT MATTER (If several classification symbols apply, indicate all)⁶

According to International Patent Classification (IPC) or to both National Classification and IPC
 Int.Cl. 5 C08F210/12; C08F2/00

II. FIELDS SEARCHED

Minimum Documentation Searched⁷

Classification System	Classification Symbols
Int.Cl. 5	C08F

Documentation Searched other than Minimum Documentation
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III. DOCUMENTS CONSIDERED TO BE RELEVANT⁹

Category ¹⁰	Citation of Document, ¹¹ with indication, where appropriate, of the relevant passages ¹²	Relevant to Claim No. ¹³
X	EP,A,0 209 253 (ENICHEM ELASTOMERI SRL) 21 January 1987 see the whole document ---	1,3,4,6
A	GB,A,561 324 (STANDARD OIL COMPANY) 15 May 1944 see claims 1-4,7 -----	

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IV. CERTIFICATION

Date of the Actual Completion of the International Search 17 AUGUST 1993	Date of Mailing of this International Search Report 01.09.93
International Searching Authority EUROPEAN PATENT OFFICE	Signature of Authorized Officer FISCHER B.R.

ANNEX TO THE INTERNATIONAL SEARCH REPORT
ON INTERNATIONAL PATENT APPLICATION NO.

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SA 73789

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Patent document cited in search report	Publication date	Patent family member(s)		Publication date
EP-A-0209253	21-01-87	CA-A-	1271896	17-07-90
		US-A-	4714747	22-12-87
GB-A-561324	None			

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